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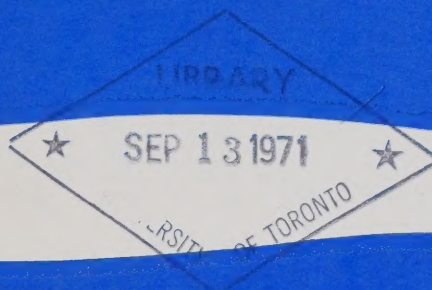
water desalination

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PREFACE

Most of us who are interested in some phase of water resource development are aware of the growing interest in water desalination. We have probably heard of the desalting plants now in operation in far-off Kuwait and Israel and know something of the intensive research activity of the Office of Saline Water in the United States. What may not be fully appreciated, however, is the likelihood that desalination may have a significant part to play in solving water supply problems not only in the extremely arid regions of the world, but also in those more humid areas where fresh water from conventional sources will be unable to satisfy all the future demands. Nor is it only in coastal regions, where there is an unlimited supply of sea water at the doorstep, that desalting will be of interest. Some communities far removed from the sea may find that extraction of smaller amounts of salt from brackish waters close at hand will prove to be an economic approach to their water supply problems. And looking into the not too distant future, we can anticipate that a growing need for careful conservation of fresh water supplies, especially in the heavily industrialized zones, will lead to the reclamation of waste water for re-use. In the reclamation process, desalination will be an important step.

These and many other technical and economic aspects of water desalination were brought to the attention of delegates from sixty different nations who met in Washington, D.C. in the fall of 1965 to attend the first International Symposium on Water Desalination. Over one hundred papers were presented to the conference by representatives of no less than eighteen different countries. They left no room for doubt about the widespread interest in desalting and the rapid progress that is being made toward development of economic processes.

The Canadian delegation returned with the firm conviction that we in Canada, despite the relative abundance of our fresh water supplies, would be well advised to keep abreast of developments in the field. Opportunities may soon arise for economic applications within our own borders. Of no less importance may be the opportunities for profitable participation by Canadian industry in the manufacture of desalting equipment.

This publication does not attempt to present an exhaustive study of water desalination, an exceedingly complex field which calls upon the expertise of the metallurgist, the physicist, the organic chemist, the biological chemist, the engineer and the economist. It is intended rather as an introduction to the subject and a review of progress in water desalination research and application. It is the hope of the authors that the publication will help the reader to place the field of water desalination in its proper perspective in the broader field of water resource development.

WATER DESALINATION

INTRODUCTION

The sparse populations and the relative under-development of the arid regions of the world are attri-butable in part to a chronic lack of adequate supplies of fresh water. In the more humid regions, populations have been expanding rapidly, accompanied in some cases by rapid industrial growth. In these more humid regions, the availability of fresh water has usually been taken for granted and scant attention paid to conservation of existing supplies. Water consumption increases with expanding populations and rising living standards, placing continuously increasing demands on fresh water resources. In the past, the demands of population and industrial centres have usually been met by reaching ever further afield for new sources of supply. The consequent extension and ramification of networks of pipelines, canals, impoundments and control works has led to progressive increases in the unit cost of fresh water.

Not only the increasing cost but even the availa-bility of fresh water, in quantity sufficient to meet future requirements, is now becoming a matter of anx-iety in many countries. More efficient use of existing supplies is becoming mandatory and the progressive adoption of such measures as pollution control, re-circulation and purification may be anticipated. There are, however, many countries where fresh water is already in short supply or where, even with careful husbanding of the existing supplies, growing popula-tions and expanding economies threaten to bring about that condition. If living standards are to be raised in countries with arid climates and if present growth rates are to be supported in those industrialized coun-tries where the situation is becoming critical, new sources of fresh water must be found. Two sources, already being exploited on a small scale, which prom-ise to play a major role in future water resource planning are sea water on the one hand and brackish water from wells on the other. It is within this back-ground that intensive research is now being carried out into the technology and economics of desalination processes.

Supporting less than one percent of the world's population and possessing over one-quarter of the world's fresh water within its borders, Canada finds itself in an enviable position compared to most other countries. Having been bountifully endowed with fresh water, Canadians might appear, at first glance, to have only an academic interest in desalination. Unfortuna-tely, however, neither the water nor the population of

Canada is distributed uniformly over the vast area of the country. There are examples within Canada's borders of semi-arid zones where copious supplies of fresh water are not readily available; there are other areas where concentrations of population and industry are not only drawing heavily on local supplies but are reaching out for new sources. Desalination of brackish water from wells may prove to be an economic solution to the problem of potable water for communities in the semi-arid zones. It is possible that desalination may be required to augment surface supplies in the more heavily industrialized areas. At some northern out-posts, where surface water is frozen for long seasons of the year and permafrost conditions make ground-water unavailable, small installations for desalting sea water might prove feasible. In areas of heavy consumption, desalination of waste water for re-use may well become an essential phase of water conser-vation practice. Canada thus has a vital interest in the progress of research and development in the tech-nology of desalination and must be alert to possible opportunities for practical applications to its specific problems.

The basic concepts involved in the removal of salt from water have been known for a long time, some of them since the dawn of history. Large-scale applica-tions of desalination, however, have had to wait not only for the development of broadly-based industrial complexes to support them but also until fresh water could no longer be provided more cheaply by drawing upon natural supplies. A demand for fresh water is not in itself sufficient inducement to make desalina-tion attractive and feasible. There must also be a sour-ce of saline water within reasonable distance of the point of consumption. There must be a source of energy at reasonable cost. The over-all economy must be able to pay for the product water and this implies the existence of other resources in the region whose exploitation will support the cost of desalting. Although certain fundamental principles will always be asso-ciated with the technical aspects of desalination, the interplay of technical, economic and social factors presents a complex picture when specific applications are considered.

Experience in desalting by various techniques and equipment in different physical and economic environments is accumulating rapidly. Experimental and operational installations are growing in orders of magnitude while unit costs of product water are decreasing. Estimates of unit cost for new installa-tions now in the planning or construction stage are

appreciably below anything yet obtained. It is within this context that the following brief review of the present status of research and development in the field of desalination is presented. Although authoritative sources have been used in its preparation, there can be no guarantee that some parts of it may not already be outdated by recent developments.

SALTS IN NATURAL AND PRODUCT WATERS

Some Characteristics of Salt Solutions

According to one modern definition, a salt is a chemical compound composed of a positive radical other than hydrogen (H^+) and a negative radical other than the hydroxyl radical (OH^-). Sodium chloride, the principal salt in sea water, is a combination of the positive sodium radical and the negative chloride radical. Examples of salts commonly found in wells and streams are the carbonates of calcium and magnesium and the sulphates of sodium and magnesium. There are many others. Calcium, sodium and magnesium are the positive radicals in the examples quoted and the carbonates and sulphates are the negative radicals.

Salts become ionized when in solution in water, the positive radicals becoming positive ions (cations) and the negative radicals becoming negative ions (anions). The solution becomes an electrolyte. When an electric current is caused to flow between electrodes immersed in the solution, the cations are attracted toward the negative electrode and the anions toward the positive. This well known principle is mentioned here because of its application in certain water desalination processes.

Some salts, such as the carbonates, tend to hydrolyze when dissolved in water, their ions combining with free hydrogen and hydroxyl ions in the water to form acids and bases. Their presence tends to give the water an acidic or basic characteristic. Other salts, such as the chlorides, sulphates and nitrates, do not hydrolyze. When only these salts are present, the water retains its neutral characteristics but acquires a saline taste. In any particular sample of natural water, total salt content is usually made up of fractions of both types. The proportion of hydrolyzed salts, expressed in terms of per cent of total dissolved salts, is called the alkalinity and the percentage of non-hydrolyzed salts the salinity of the sample.

It must be emphasized that desalination usually involves the removal of all types of salts, regardless of how they appear in solution. Therefore, when terms such as "saline content" or "salt content" are used in this report, they refer to the total dissolved salt, including both the hydrolyzed and non-hydrolyzed constituents.

The quantity of dissolved solids in a sample of water is expressed in "parts per million (ppm)", the weight in grams of the solids in one thousand litres of the water.

The total weight of dissolved solids in a water sample is the sum of the weights of all the anions and cations of the salts plus the weight of such other impurities as colloidal silica and ion oxide. These non-salt impurities, which are usually a relatively small proportion of the total dissolved solids in natural waters, will be removed along with the salts in most desalination processes.

The "hardness" of water, which makes it unsatisfactory for washing purposes, is due to the presence of bicarbonates of calcium or magnesium (temporary hardness) or the sulphates of calcium and magnesium (permanent hardness). Temporary hardness can be removed from the water by boiling, which tends to precipitate the salt in the form of carbonates. Permanent hardness can be removed only by distillation or other more sophisticated desalination method.

Saline Content of Source Waters

In some desalination processes, the quantity of salt to be removed has a significant effect on the unit cost of the product water. In others, the quantity of salt is not significant. For this reason, the total amount of dissolved solids in the source water has an important bearing on the economics of a specific application and on the selection of the process to be used.

Saline waters, as they occur in nature, are divided into two broad categories: sea water and brackish water. Ordinary sea water, which has an average salt content of 35,000 ppm, lies near the upper end of the scale, exceeded only in a few inland seas in extremely hot and arid regions. Brackish water is strictly defined as any water with less dissolved salts than sea water. The term is usually applied to waters containing somewhat more than 500 ppm, generally considered the maximum level for human consumption. Waters with less than 500 ppm are usually referred to as "fresh" waters, although the terms "brackish" and "fresh" are relative ones and the distinction is not rigidly defined. Brackish water may occur in nature with salt content anywhere in the range below 35,000 ppm, depending upon the source.

Excluding the semi-arid regions in the Prairie Provinces, the salt content of surface waters in Canada is generally less than 500 ppm and in many instances is below 100 ppm. On the prairies, there are many brackish lakes and streams where the salt content may exceed 1000 ppm. The quality of water from wells

varies over a wide range from fresh to extremely brackish. There are many communities on the Canadian prairies that are using well water with salts exceeding 500 ppm and a few extreme cases where the value exceeds 2000 ppm.

Permissible Saline Content of Product Waters

Since the quantity of salt to be removed in any specific application will depend not only on the amount of salt in the source water but also on the required quality of the product water, the use to be made of the desalinated water becomes another important consideration in assessing the economic feasibility of desalination and the process to be used.

A value of 500 ppm dissolved salts is considered to be the desirable maximum for human consumption, although up to four times that amount is now tolerated in some parts of the world where better water is not readily available. Domestic animals can tolerate somewhat higher saline content than human beings.

Specifications for industrial processes vary over a wide range. Sea water is feasible for some cooling purposes, despite the corrosion problem, while some industrial uses call for a considerably higher quality than is acceptable for drinking water. An extreme case on the lower end of the scale is the requirement for feed water in modern steam power generation, where even 10 ppm is excessive.

As a general principle, irrigation water should be as low in dissolved salt as possible in order to preserve soil fertility. A salt content of 1000 ppm is considered the maximum permissible for agricultural purposes, although some crops require a much higher standard and a few can be grown successfully with a more saline water. These figures are only of academic interest at this time because there is little expectation that desalination can produce water economically in sufficient quantity to support a significant agricultural operation. Exceptions to this generality may arise in the future in very arid regions, where dilution of irrigation water with product water from a desalter might prove to be profitable in raising certain high-value produce.

DESALINATION PROCESSES

Classification of Processes

Several different principles are represented among the various water desalting processes now in use or under investigation. The principle of distillation, which has been known and practised on a small scale by mankind for thousands of years, is the one that received the first attention during the early stages of

the current development program. The trend has been toward sophistication of techniques and equipment. Modern distillation processes are classified as submerged-tube distillation, flash evaporation, long-tube vertical distillation or vapor compression, depending upon the methods and equipment employed in the evaporation and condensation phases. The use of solar stills, which is a cruder method of distillation, may continue to have interest for small installations in some areas of the world.

Some processes take advantage of the filtering properties of various types of membranes, some of which are highly selective with respect to water and not salt and others to the positive or negative salt ions. Electrodialysis and reverse osmosis are the principal applications. Separation by freezing, either with or without the use of a secondary refrigerant, is another principle which is being employed in some demonstration plants. The property of certain resins to absorb salt ions is the basis of another process known as ion exchange. Other techniques involving separation by use of water solvents or hydrocarbons are under study at the laboratory level.

Desalination processes may also be classified according to the type of energy employed. Thermal energy is used for some processes, including most of the distillation techniques. Others require either electric or mechanical energy. A few require no energy input other than that needed for pumping purposes. None of the methods with practical application at the present time is in this latter category.

Processes also fall into two broad classes depending upon the type of source water to which they are applicable. Those in which energy consumption is not influenced appreciably by salinity may be used for treatment of water of any salt content, including sea water. The various distillation and freezing techniques are of this type. In some other processes, electrodialysis being the most important example, energy consumption is directly proportional to the amount of salt to be removed from each gallon of the source water. These processes may have advantages in the treatment of certain brackish waters but cannot compete where desalting of sea water is concerned.

Processes may be further classified with respect to the maximum feasible capacity of the desalting units. It now appears that engineering problems will place a severe restriction on this capacity in some processes. Capital costs in these cases are likely to increase directly with the size of the plant. In other processes, the maximum feasible capacities of the individual units are not clearly discernible at this time but will be appreciably higher than in any existing plants. Such processes will acquire a distinct advantage for future large-scale developments.

TABLE 1

Desalination Processes in Use or Under Investigation

Principle	Process	Source Water	Energy Required (excluding pumping)	Phase Changes
Distillation	Multi-effect submerged-tube distillation	Sea water or brackish water	Thermal	Liquid to vapor to liquid
	Flash evaporation (a) Multi-stage (MS) (b) Multi-effect, multi-stage (MEMS) (c) Vapor-reheat	Sea water or brackish water	Thermal	Liquid to vapor to liquid
	Long-tube vertical distillation (LTV)	Sea water or brackish water	Thermal	Liquid to vapor to liquid
	Vapor compression	Sea water or brackish water	Mechanical and some thermal	Liquid to vapor to liquid
	Solar distillation	Sea water or brackish water	Solar	Liquid to vapor to liquid
Separation using Membranes	Electrodialysis	Brackish water	Electric	None
	Reverse Osmosis	Sea water or brackish water	Mechanical	None
Freezing	Direct Freezing	Sea water or brackish water	Mechanical	Liquid to ice crystals to liquid
	Indirect freezing — using a refrigerant	Sea water or brackish water	Mechanical	Liquid to ice crystals to liquid
Ion Exchange	Ion exchange	Brackish water	None	None
Separation using Hydrating Agents	Hydrate process	Sea water or brackish water	Mechanical	Liquid to hydrate crystals to liquid
Separation using Solvents	Solvent extraction (or organic absorption)	Sea water or brackish water	Thermal	None; water to water-in-solution to water.

A further discrimination may be made between processes which have already been successfully demonstrated as feasible at the plant level and those which are still in the laboratory and whose future is still in doubt. Electrodialysis and various distillation processes are in the first category and reverse osmosis and freezing techniques may also be included. None of the other possible alternatives can be considered to have immediate practical significance.

The desalting processes now in use or under study are listed in Table 1. The kind of source water to which the process is applicable and the type of input energy required are indicated in each case.

Distillation Processes

MULTI-EFFECT, SUBMERGED-TUBE DISTILLATION:

In the most simple application of desalination of water by distillation, heat is applied to saline water until evaporation takes place. The vapor is drawn off and condensed to produce the required distillate, the remaining water in the evaporating vessel becoming progressively more saline. Recent research has been directed at the improvement in application of this principle to produce maximum thermal efficiency in high-capacity installations at minimum capital cost.

Modern distillation practice had its beginnings near the end of the nineteenth century in marine installations providing feed water for boilers in ocean vessels. Early applications were of the "single-effect" type in which sea water was boiled in a vessel, heat being supplied by steam in tubes immersed in the brine. Pre-heating of the brine was achieved by condensing the product vapor around coils containing the incoming brine. The search for higher thermal efficiency in larger plants led to the development of "multi-effect" systems, in which the brine was passed from boiler to boiler, each boiler being operated at lower temperature and pressure than the preceding one (Figure 1). In such systems, the input heat to the first boiler induces evaporation and the resulting vapor is led into condensing coils immersed in the brine in the next boiler or "effect". Condensation of the vapour causes the hot brine in the second effect to boil. The vapor from this effect is then condensed
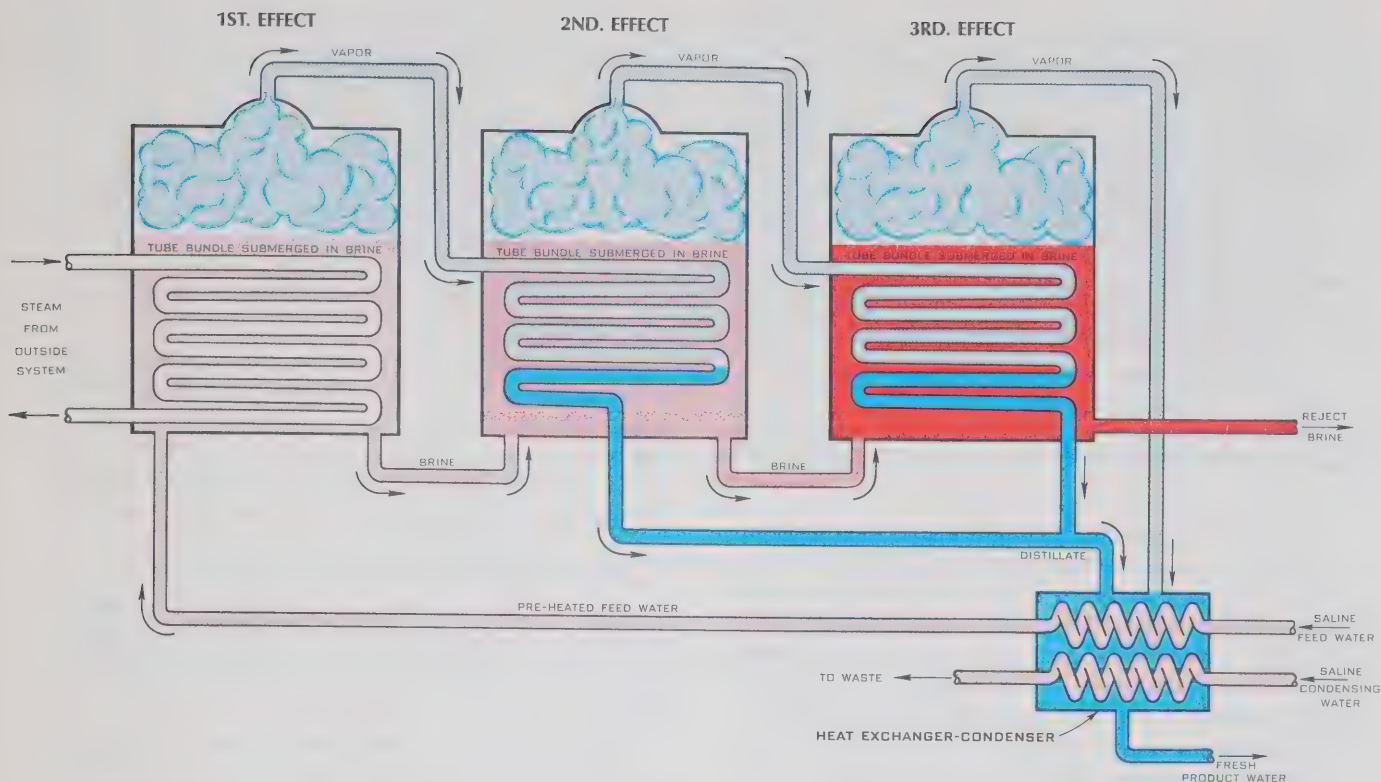
FIGURE 1. MULTI-EFFECT, SUBMERGED-TUBE DISTILLATION

in the third effect, and so on. The first effect has a single function as the primary boiler, succeeding effects act as both boiler and condenser and the last in line may perform the single function of a condenser. The condensate is drawn off from each of the condensing coils as product water. The highly concentrated high-temperature brine from the last effect is used to pre-heat the feed water to the first effect and is then rejected.

The multi-effect submerged-tube process evolved through the first half of the twentieth century, reaching its most advanced stage around 1958 in six-effect units of about 0.5 million U.S. gpd capacity installed in Aruba and Curaçao in the Netherlands Antilles. At this point it was becoming apparent that improvements in efficiency and economy in larger installations of this type would be inhibited by design and engineering problems. The process employs a complex series of heat exchanges of both the "liquid-inside" type (in pre-heating of the feed water) and "liquid-outside" type (in the evaporator-condenser units) and cost of the equipment becomes critical as capacity is increased. The immersion of more tubes requires greater depths of brine, resulting in increased hydrostatic head and higher boiling point in the evaporators at a cost in thermal efficiency. Furthermore, high maintenance costs in combatting scale and corrosion cannot

be avoided when saline water is boiled in contact with metal heat-exchange surfaces. These drawbacks led to development of alternative processes and it now appears that submerged-tube distillation will not play a significant part in large-scale water desalination in the future.

MULTI-STAGE FLASH EVAPORATION:

In the search for more economic methods, a logical step beyond submerged-tube distillation was the multi-stage, flash evaporation process. As feed water passes through the successive effects in a submerged-tube system, some of it tends to flash directly into vapor as it enters each effect. The flash evaporation process, by inducing all the vaporization to take place by flashing, dispenses with the boiling operation and thus eliminates the bundle of submerged tubes at significant saving in capital and maintenance costs.

In practical applications of the flash evaporation technique, the saline feed water is passed through a series of evaporation chambers or "stages" under carefully controlled pressure and temperature conditions (Figure 2). Pressure is reduced progressively through the system in such a way that the pressure in each stage is a little below the vapor pressure of the incoming brine. Part of the brine flashes into vapor

immediately upon entrance and the remainder is passed to the succeeding stage. Vapor is drawn off from the evaporating areas and is condensed around tubes, the condensate being collected on trays as product water. Heat recovery is effected by passing incoming feed water through the series of condenser tubes in the opposite direction to the flow of the brine through the evaporators. After emerging from the first-stage condenser, the pre-heated feed water is raised to boiling temperature in a heat exchange; using process steam or other outside heat. It is then injected into the first-stage evaporator. Part of the concentrated brine emerging from the last stage is re-circulated by mixing with the feed water. The remainder of the concentrated brine is rejected.

The first important industrial application of multi-stage, flash evaporation (multi-flash) was a 4-stage, 4-unit plant of 2 million U.S. gpd total capacity, which went into operation in Kuwait in 1957. Multi-flash has now replaced the submerged-tube process and at the present time is still the favored process where units exceeding 100,000 U.S. gpd are required. Its principal advantages over earlier methods are the reduction in capital costs by the elimination of the submerged tubes and the lower maintenance costs achieved by elimination of scale and corrosion problems associated with the boiling of brine in contact with metal surfaces. On the other hand, the process still requires large banks of tubes for condensing and

feed water pre-heating. The cost of the tubes may represent over one-half the capital cost or over one-third the total cost of the product water. For any given number of stages, the number of tubes tends to increase directly with capacity. For this reason, expansion of units beyond the range of existing plants (1,500,000 U.S. gpd in Curaçao) is unlikely, in itself, to effect major reductions in the costs of product water.

MULTI-EFFECT, MULTI-STAGE FLASH EVAPORATION:

Recent research has led to the combination of some of the principles of multi-effect distillation and multi-flash evaporation in a process known as multi-effect, multi-stage flash evaporation (MEMS). In the previously-described single-effect, multi-stage process (SEMS), the preheating of the feed water is completed before it enters the first-stage evaporator. At this point in the system, the pressure differential required to pass the brine through the successive stages is already established and a practical limitation is imposed on the number of stages and the rate of flow of the brine.

An MEMS system employs a series of effects (Figure 3), each of which consists of a number of stages operated in much the same way as in an SEMS system. Brine from the lower end of each effect passes to the first stage of the next effect. Vapor is drawn off from each effect and condensed around coils containing the counter-flowing feed water, just as in a
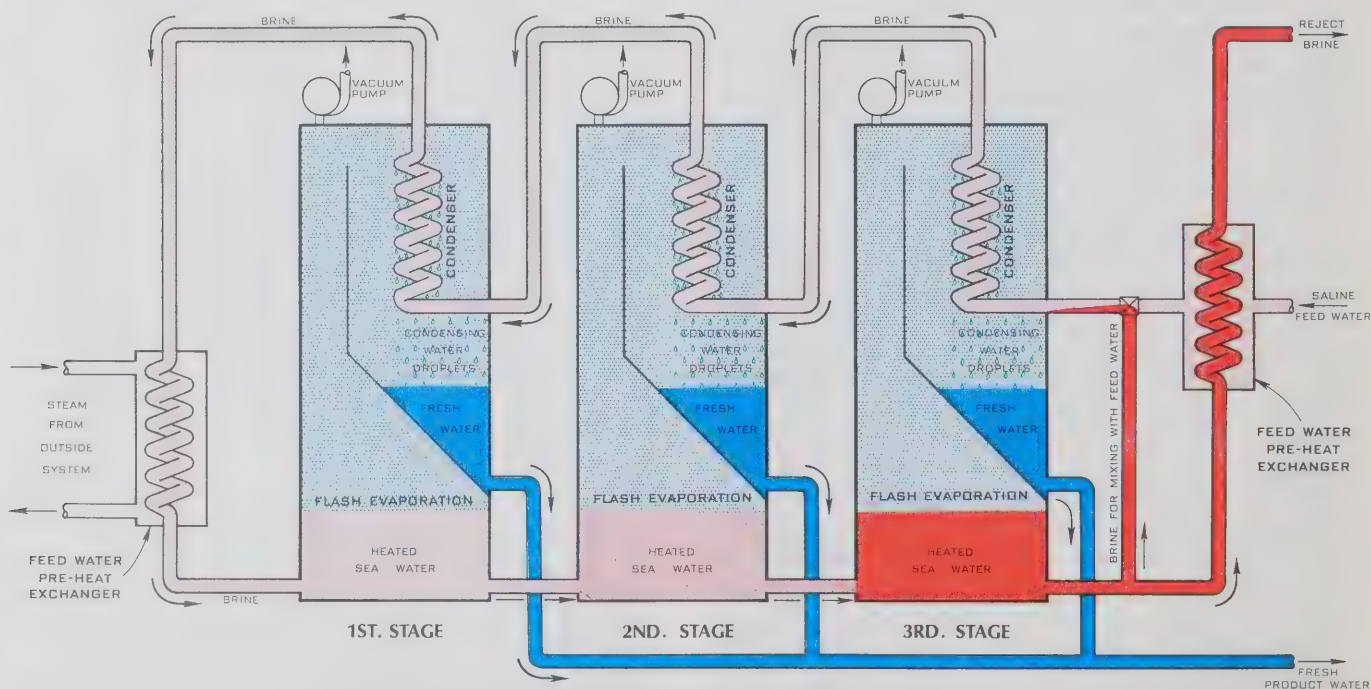
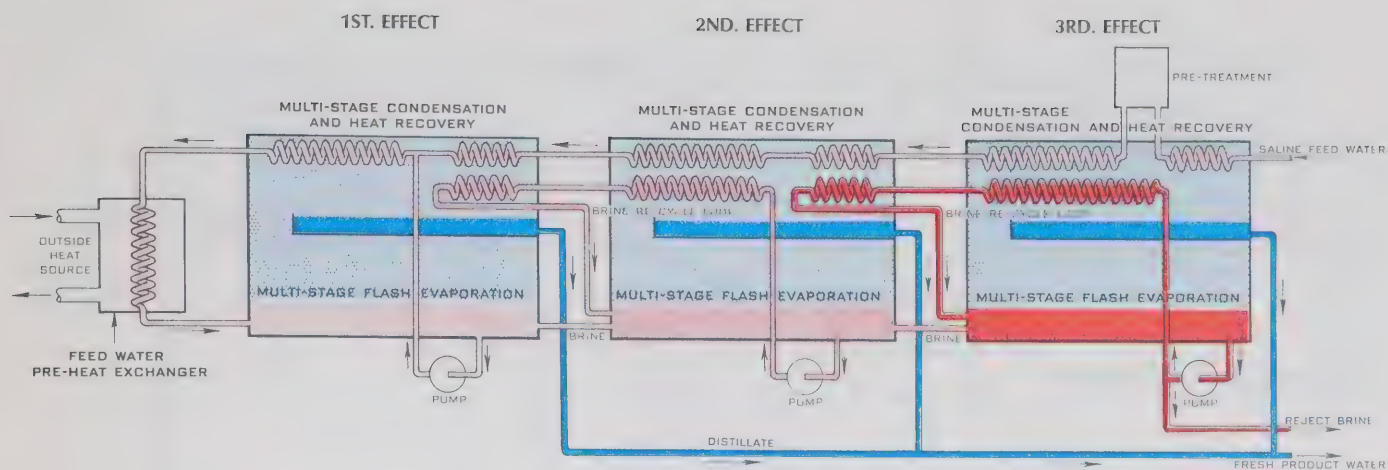FIGURE 2. MULTI-STAGE FLASH EVAPORATION

FIGURE 3. MULTI-EFFECT, MULTI-STAGE FLASH EVAPORATION

typical SEMS system. The distinguishing feature in the MEMS system is the use of part of the vapor from the last few stages in each effect to raise the temperature of the brine in the succeeding effect. This is accomplished by introducing secondary loops in which some of the brine from the second and succeeding effects is re-cycled through condensing coils in the last few stages of the preceding effect, after which it is re-injected into the top stage of the effect from which it was drawn. Adding the re-cycling loops to the single loop of the SEMS system improves economy and thermal efficiency by raising the rate of circulation of brine in the high-temperature effects and by increasing the number of stages that can be used in the system.

The MEMS method has been successfully demonstrated in a 4-effect, 64-stage, 7,200 U.S. gpd test plant developed by the American Machine and Foundry Company of Waterford, Connecticut. A 3-effect, 68-stage plant is proposed for installation in San Diego as a replacement for the SEMS demonstration plant which was recently moved to Guantanamo Bay in Cuba.

VAPOR-REHEAT FLASH EVAPORATION:

Now under study is a suggested modification of the multi-flash process, known as the vapor re-heat process, which may make it possible to dispense with the tube banks entirely. In this process, vapor produced by multi-stage flashing would be drawn off from each stage and condensed on the surface of fresh water which is being re-cycled through tanks in a direction counter to the flow of brine (Figure 4). The fresh water stream, emerging at high temperature from the top stage, would be used to pre-heat incoming feed water and then re-cycled through the condensing tanks, beginning at the bottom stage. Product water in quantity equal to the condensate would be drawn off from
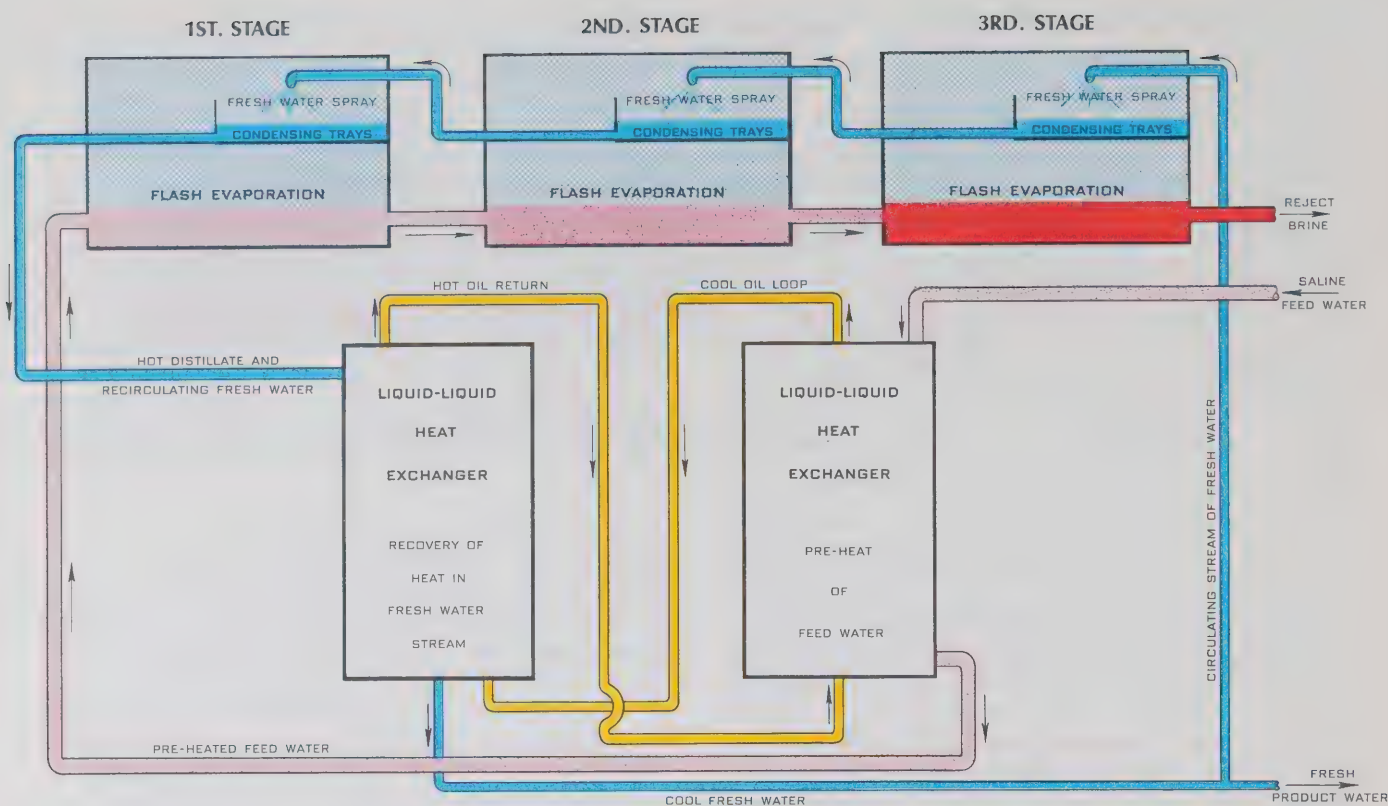
the fresh water stream after it has been cooled in preheating the feed water and before it is re-cycled.

The process, as described, would dispense with the condenser tubes of the conventional multi-flash process but would still employ a heat exchanger with metallic heat-transfer surfaces in the feed water preheater. A possible further modification would replace this metal heat exchanger with a "liquid to liquid to liquid" exchanger, thus eliminating tubes or other metallic heat-transfer surfaces entirely. In such an exchanger, a non-miscible liquid such as kerosene would be decanted with the stream of hot, fresh water from the condensing tanks. The kerosene, after picking up heat from the fresh water, would be passed to another decanter, where transfer of heat to the feed water would be effected. The kerosene would be continuously re-cycled through the two decanters.

LONG-TUBE VERTICAL DISTILLATION:

A process which has been used for many years in the chemical industry, long-tube vertical distillation (LTV), has been under investigation since 1955 as an alternative for large-scale water desalination. It is actually a sophistication of the submerged-tube process in that vaporization is produced by boiling the feed water in a series of effects, the vapors produced in each effect being condensed in the next, where the heat recovery induces further boiling (Figure 5). The principal difference in the LTV process is that the feed water is boiled as it falls through banks of vertical tubes and the necessity for submergence of the tubes in the brine is eliminated.

In practical applications of the LTV process, the saline water is introduced to a header tank at the top of the first effect from where it falls through the vertical tubes, the space around the tube bundle

FRESH WATER SPRAY
CONDENSING TRAYS
FLASH EVAPORATION

FRESH WATER SPRAY
CONDENSING TRAYS
FLASH EVAPORATION

FRESH WATER SPRAY
CONDENSING TRAYS
FLASH EVAPORATION

REJECT BRINE

SALINE FEED WATER

HOT OIL RETURN COOL OIL LOOP

HOT DISTILLATE AND
RECIRCULATING FRESH WATER

LIQUID-LIQUID

HEAT

EXCHANGER

RECOVERY OF

HEAT IN

FRESH WATER

STREAM

LIQUID-LIQUID

HEAT

EXCHANGER

PRE-HEAT

OF

FEED WATER

CIRCULATING STREAM OF FRESH WATER

PRE-HEATED FEED WATER

COOL FRESH WATER

FRESH PRODUCT WATER

NOTE: HEAT EXCHANGERS WITH METAL TRANSFER SURFACES MAY BE SUBSTITUTED
FOR THE LIQUID-LIQUID EXCHANGERS.

FIGURE 4. VAPOR-REHEAT FLASH EVAPORATION

being filled with steam from outside the system. The water falls in a film, clinging to the inside walls of the tubes, part of it boiling during the descent. A mixture of vapor and brine emerges from the bottom of the tubes and is collected in a chamber from which the vapor is drawn off and introduced to the jacket around the tube bundle of the second effect. Meanwhile the brine remaining from the first pass is pumped to the top of the second effect, where it again falls through vertical tubes. In this effect, the vapor from the first effect condenses on the outside of the vertical tubes to form the product water, at the same time inducing boiling in the film of brine on the inside walls of the tubes. The process is repeated in succeeding effects, product water being drawn off from the outside of the tube bundle in each effect. Some of the brine emerging from the tubes in the last effect is mixed with incoming feed water and re-cycled. The remainder of the brine is rejected. Residual heat in the various effluents is recovered by suitable heat exchangers.

The LTV process eliminates the engineering and design problems connected with the submergence of

tubes but does not eliminate the problems associated with the boiling of brine in contact with the walls of the metal tubes. The economic feasibility of the method is governed to a large extent, therefore, by the measure of success in control of scale and corrosion. Pretreatment of feed water to control pH factor is one approach . Another approach is the use of the slurry method which involves seeding the feed water with scale crystals to encourage precipitation of the scale around the crystals instead of on the tube walls. Research is being directed also toward the selection of the most corrosion-resistant alloys for the heat-exchangers and other fittings.

The largest application of the LTV process now producing desalted water is the 12-effect, 1 million U.S. gpd demonstration plant in Freeport, Texas, developed by the Office of Saline Water of the U.S. Department of the Interior. Sea water of 35,000 ppm salinity is used to produce fresh water with 50 ppm maximum dissolved solids. Part of the product water is being used industrially and the rest to augment the Freeport municipal supply.
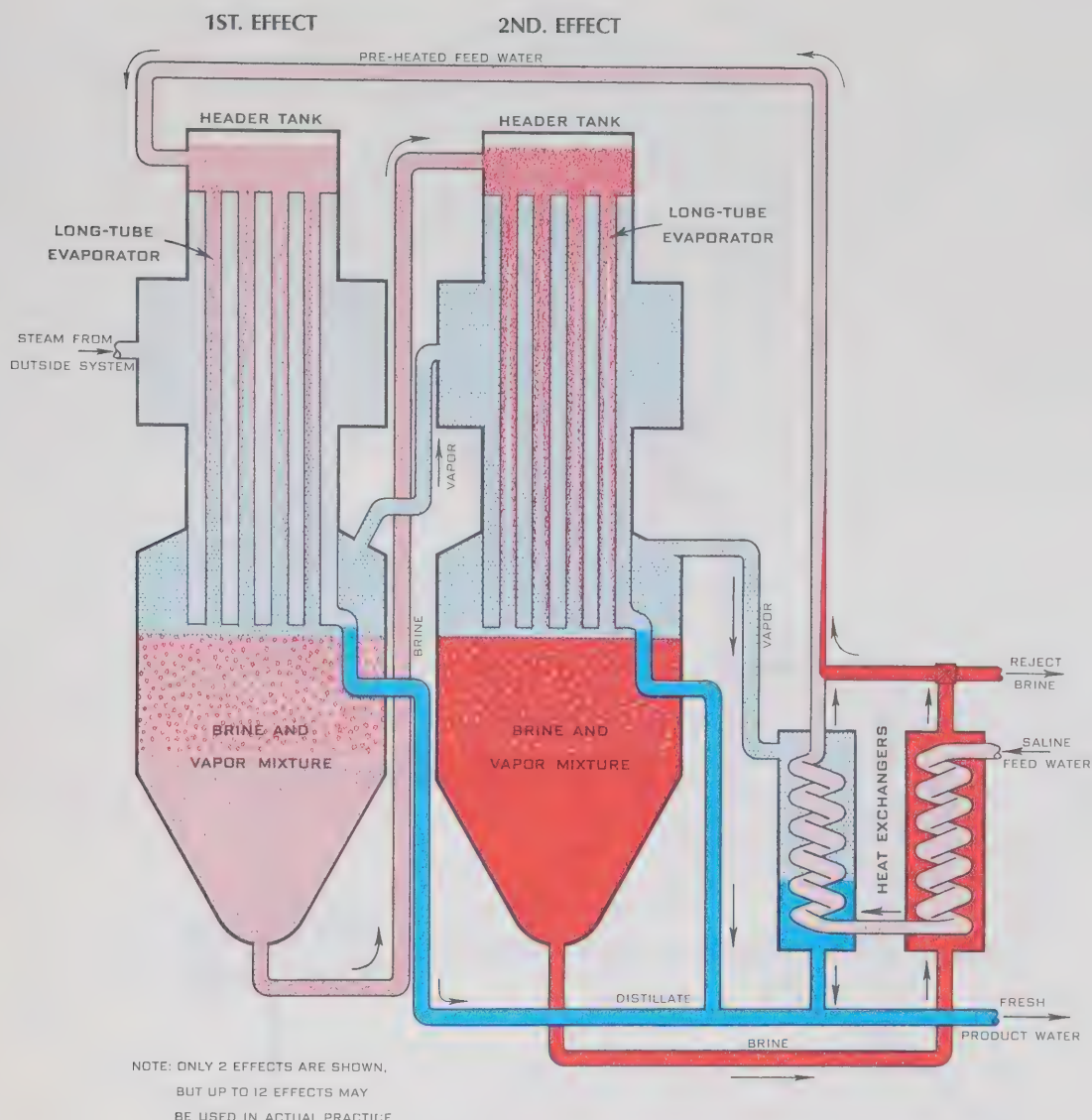
1ST. EFFECT **2ND. EFFECT**

PRE-HEATED FEED WATER

HEADER TANK

HEADER TANK

LONG-TUBE
EVAPORATOR

LONG-TUBE
EVAPORATOR

STEAM FROM
OUTSIDE SYSTEM

VAPOR

BRINE

BRINE AND
VAPOR MIXTURE

BRINE AND
VAPOR MIXTURE

VAPOR

VAPOR

REJECT
BRINE

HEAT EXCHANGERS

SALINE
FEED WATER

DISTILLATE

FRESH
PRODUCT WATER

BRINE

NOTE: ONLY 2 EFFECTS ARE SHOWN,
BUT UP TO 12 EFFECTS MAY
BE USED IN ACTUAL PRACTICE.

FIGURE 5. LONG-TUBE VERTICAL DISTILLATION

VAPOR-COMPRESSION DISTILLATION:

Vapor compression is another approach to distilla-
tion which has been successfully demonstrated. In
the basic process, steam is drawn from an evaporating
chamber and its temperature and pressure raised in a
vapor compressor. It is then passed to a jacket surround-
ing a vertical tube bundle where it is condensed on the
outside of vertical tubes containing the saline feed
water (Figure 6). The feed water is introduced to the
bottom of the tubes and is pumped upward through
them, reaching boiling temperature as it picks up heat
from the condensation of the vapor in the surrounding
jacket. As it emerges from the tops of the tubes, part
of the saline water flashes into vapor, which is then
passed to the compressor to complete the cycle. The
unevaporated portion of the saline water is drawn off

the bottom of the vapor dome as hot, concentrated
brine. Part of it is re-circulated by mixing with the
incoming feed water and the remainder is rejected,
after passing through a heat exchanger where residual
heat is given up to the feed water.

Almost all the input energy to the system is from
the mechanical energy of the compressor, only a very
small make-up of heat being required to offset the
minor heat losses. It is on this account that the proc-
ess may be of particular interest to areas where hydro-
electric energy is available to drive the compressor
and where thermal energy from fossil fuels is relatively
expensive.

The process may be used for both sea water and
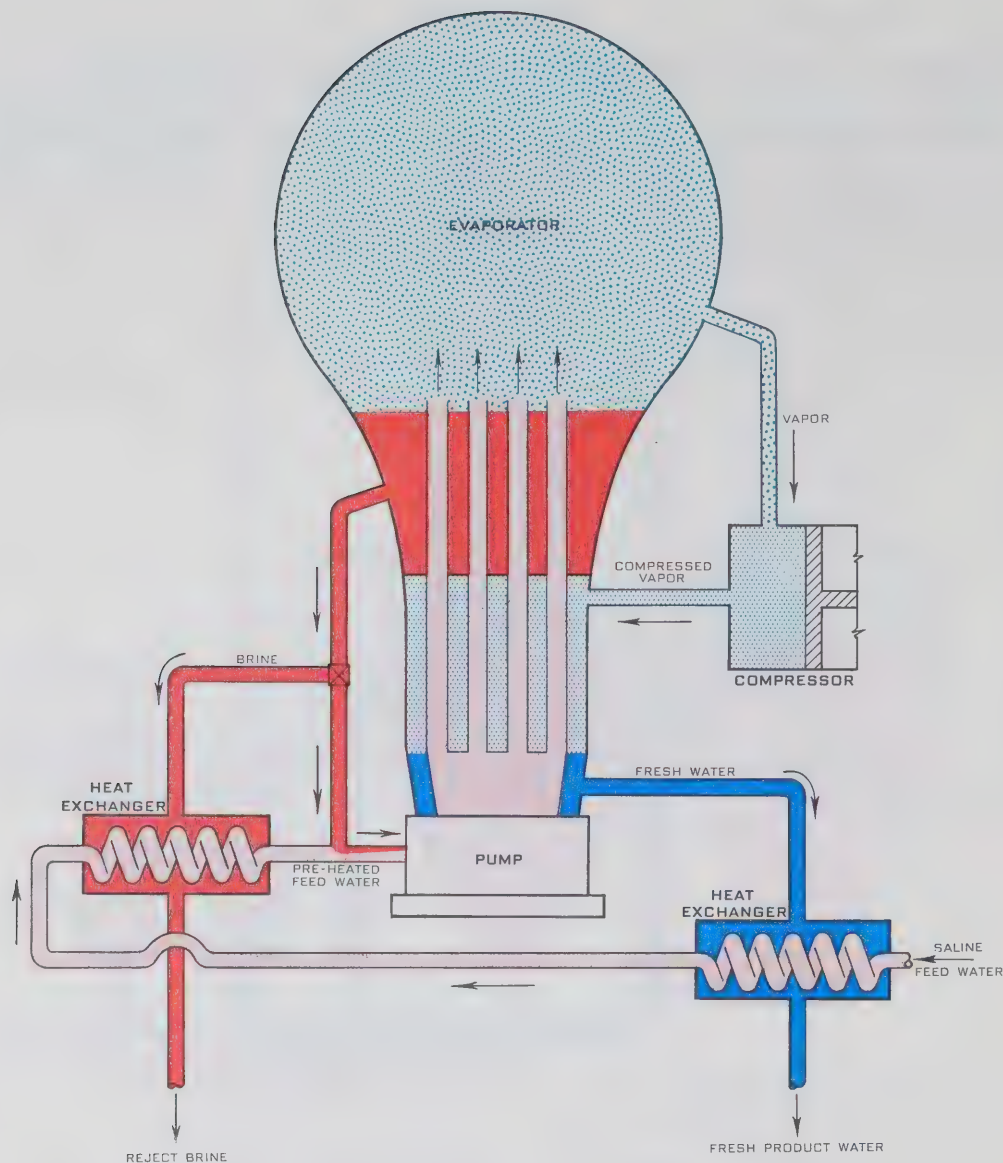brackish water and can be operated successfully

FIGURE 6. VAPOR-COMPRESSION DISTILLATION

where the source water has a relatively high content of scale-forming salts. It requires no cooling water. These features make it attractive in those arid regions where the only source available is brackish water in small quantities. Its principal disadvantages are the relatively high capital cost and the limitation on the size of individual units imposed by engineering problems.

The outstanding example of a practical application of the process at the present time is a plant of one million U.S. gpd capacity at Roswell, New Mexico, operated by the U.S. Office of Saline Water. It is a demonstration plant which is producing fresh water to augment municipal supply for the city of Roswell. In this case, the basic process has been extended to a 2-effect system in which the vapor from the first effect is passed directly, without compression, to the outside of the tube bundle in the second effect. The vapor from the second effect is compressed and then injected into the first effect to complete the cycle. The unevaporated brine from the first effect is passed through the tubes in the second effect where further evaporation results in a residual brine of about four times the salinity of the original feed water. Both the product water from the outside of the tube bundles and the brine remaining from the second effect are passed through heat exchangers, where their residual heat is used to raise the temperature of the feed water to 224°F before its injection into the first evaporator.

Feed water, which is extremely brackish, is given extensive treatment by ion exchange, acid treatment and de-aeration before introduction to the evaporators.

SOLAR DISTILLATION:

Although the use of radiant energy of the sun in the operation of simple stills is man's oldest method of water desalination, it will have no application to modern large-scale plants unless and until a major technological breakthrough occurs, principally because the unit cost of the product water is relatively high, despite the fact that the input energy is free. Small solar stills can find a place in special circumstances in hot and arid regions, however, where the successful exploitation of other resources would be inhibited by lack of suitable drinking water.

In the pilot installations now under study, the saline water is exposed to the sun in shallow, concrete pans with blackened floors (Figure 7). The vapor rises from the water surface by convection to condense on the under side of a roof of transparent glass or plastic, which is supported close above the pans and extends over the entire area of the still. The roof is constructed of individual sheets, inclined at an angle and joined so as to give a corrugated appearance. The condensate runs down the under side of the sheets and drips into gutters under the troughs of the corru-

gations. It is drained off from the troughs as product water.

Maximum capacity so far achieved is about 1 U.S. gpd per 10 square feet of water surface, which represents about 50% thermal efficiency. One of the problems is the tendency for dust to collect on the condensing surfaces. The dust absorbs some of the radiation, reducing the amount reaching the water and lowering the cooling capacity of the condensing surfaces.

Various modifications in the design of the stills are being considered. It has been suggested, for example, that the economics of solar distillation in some tropical areas could be improved if the stills were designed to fulfil a double function, acting as conventional stills when the sun shines and as catchments during periods of rain.

Processes Using Membranes

ELECTRODIALYSIS:

Membranes of synthetic resin are available which are highly permeable to the anions of a saline solution but not to the cations. There are others that have the opposite characteristics. The selective properties of these membranes are used to advantage in a filtering technique for desalting known as electrodialysis.
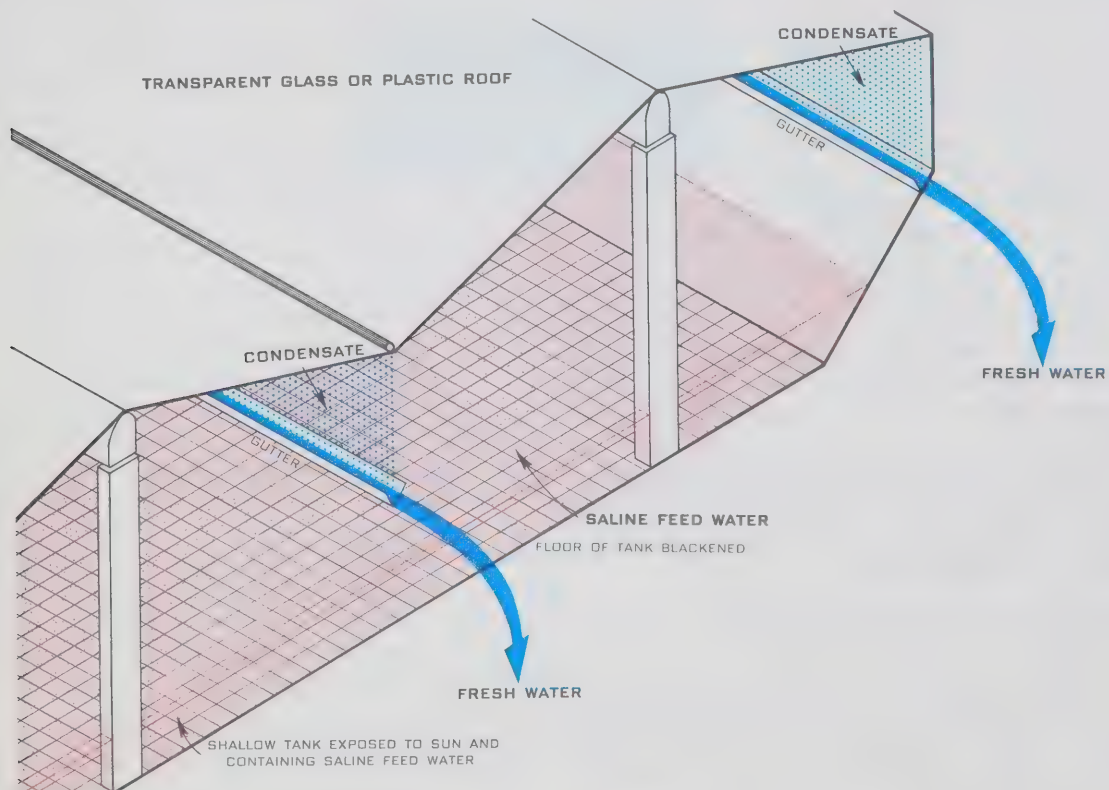
TRANSPARENT GLASS OR PLASTIC ROOF

CONDENSATE

GUTTER

FRESH WATER

CONDENSATE

GUTTER

SALINE FEED WATER

FLOOR OF TANK BLACKENED

FRESH WATER

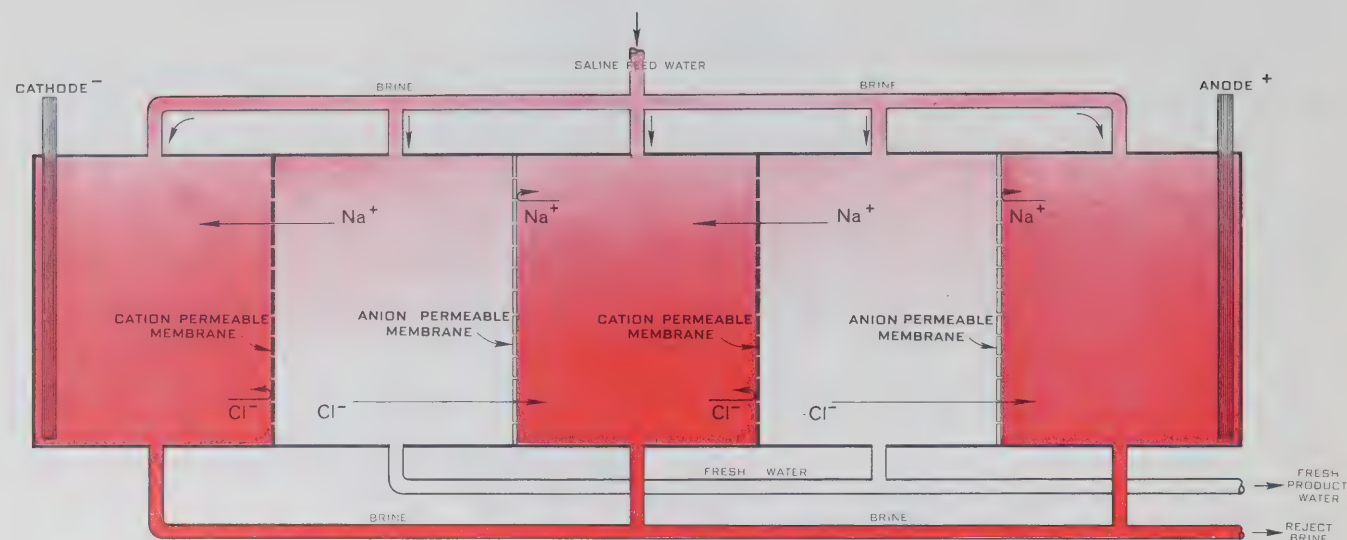SHALLOW TANK EXPOSED TO SUN AND CONTAINING SALINE FEED WATER

FIGURE 7. SOLAR STILL

11

FIGURE 8. ELECTRODIALYSIS

A typical electrodialysis unit is composed of a "sandwich" of hundreds of membranes, arranged alternately with respect to their proclivity to the passage of ions (Figure 8). Feed water is introduced to one end of the unit or "stack" and flows between the membranes in the cells formed by the membranes and the frame of the stack. A direct current is passed through the water in a direction perpendicular to the direction of flow of the water, attracting the anions of the dissolved salts toward one side of the stack and the cations toward the opposite side. The migrating ions pass through the membranes enclosing every alternate cell, the anions emerging from one side and the cations from the other side. Meanwhile, the ions are accumulating in the intervening cells, their further lateral migration obstructed by a membrane of the type through which they are unable to pass. Thus, during its passage through the stack, the feed water becomes progressively more saline in every alternate cell and more dilute in the intervening ones. At the outlet end of the stack, streams are drawn off from the individual cells either as the product water or brine for rejection, as the case may be.

In larger electrodialysis installations, the multi-stage principle is employed by linking two or more stacks in series. The dilute effluent from the first stack or "stage" is passed through an identical second stage, and so forth, the effluent from the final stage reaching the dilution required for the product water.

The process consumes electric power in direct proportion to the amount of salt to be removed from each unit of feed water. Consequently, the cost of operation is governed by both the salt content of the source water and the specifications for the product water. For this reason, the process is not favored for treatment of sea water or for production of high-quality fresh water. It is particularly adaptable to situations requiring reduction of dissolved salts in brackish water from about 5000 ppm to about 500 ppm. It may well play an important role in the future in the reclamation of sewage, where desalination will be one of the essential phases.

Energy consumption in an electrodialysis process decreases with the increase in water temperature but the advantage tends to be offset by more rapid deterioration of the membranes. Energy consumption increases with deposition of scale on the membranes, the quantity of scale being related to the types of dissolved salts. To combat this scale problem, pre-treatment of the feed water is usually required. As brackish waters, unlike sea water, vary widely in the quantity and nature of their dissolved salts, the economic and technical aspects of any particular application will be governed by the composition of the source water to be treated. A problem common to all, and one which is receiving intensive research, is the tendency to high cost and short life of the membranes.

The largest existing electrodialysis plant is a 2-stage development in South Africa with a capacity approaching 3 million U.S. gpd, where 500 ppm water can be produced from brackish water of 3000 ppm salt content. The plant went on stream in 1959 but is now on

stand-by status. Technical problems with the membranes developed but it is reported that they have now been largely resolved.

The U.S. Office of Saline Water is operating a demonstration plant at Webster, South Dakota. It is a 4-stage process with a one-quarter million U.S. gpd capacity in which a salt content of 1800 ppm in the brackish source water is reduced to 275 ppm in the product water. There are numerous small installations throughout North Africa and Asia Minor with capacities ranging downward from about 100,000 U.S. gpd, where fresh water of about 500 ppm is produced from brackish source waters with a salt content in the order of 5000 ppm. Another significant installation is a 650,000 U.S. gpd plant at Buckeye, Arizona, where a 500 ppm water is produced from brackish well water with about 2100 ppm dissolved salts. The plant is owned and operated by the municipality and supplies the entire municipal demand.

REVERSE OSMOSIS:

When a body of pure water and a salt solution are separated by a particular kind of membrane which is permeable to water but not to ions of dissolved salt, some of the pure water flows through the membrane to dilute the salt solution. Osmosis, as the process is called, continues until the pressure in the salt solution exceeds the pressure in the pure water by an amount called the osmotic pressure of the solution. Sea water has an osmotic pressure of 350 psi and brackish waters a lesser value, depending upon salt content.

When a pressure in excess of its osmotic pressure is applied to the salt solution in the circumstances described, the direction of flow of the water is reversed and the salt solution becomes progressively more concentrated while the quantity of pure water on the opposite side of the membrane increases. This reversal of the normal osmosis process is another possible method for producing fresh water from a saline source (Figure 9).

Research on reverse osmosis has reached the point where it has been successfully demonstrated in a pilot plant of 1000 U.S. gpd capacity. Product water of 500 ppm has been obtained from sea water and product water of 250 ppm from brackish water with 4500 ppm
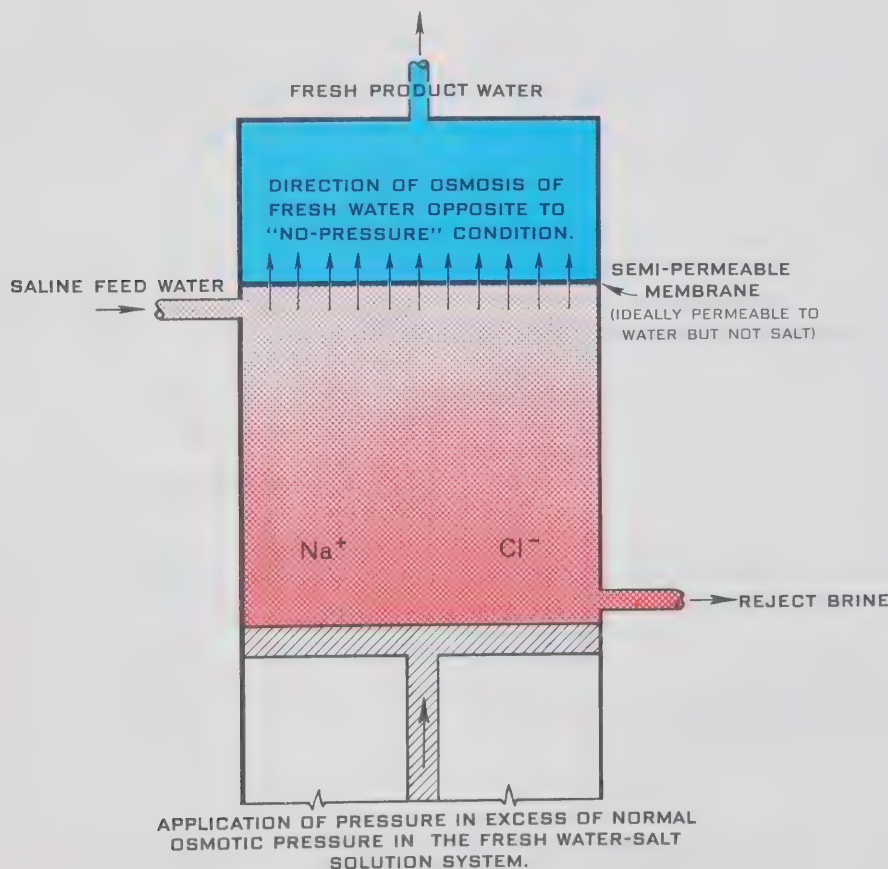

FIGURE 9. REVERSE OSMOSIS

dissolved salts. Experience with the plant has demonstrated that the flux for any particular membrane varies directly with the effective pressure (the difference between the operating pressure and the osmotic pressure of the source water) and that salt penetration of the membrane varies inversely with the effective pressure. A higher pressure thus results in a higher quality in the product water. Operating pressures up to 1500 psi for sea water and up to 750 psi for the brackish water have been used in the pilot plant.

An even larger pilot plant at Coalinga, California, designed to produce 5000 U.S. gpd of 300 ppm water from a brackish feed water of 2500 ppm, has been under test since June 1965. Difficulty has been experienced in reaching design capacity, however, because of deposition of slime on the membranes. The product water is being used to augment the town water supply.

The U.S. Office of Saline Water has announced that contracts have been awarded for construction of two pilot plants in California, one of 50,000 U.S. gpd capacity and the other a 10,000 U.S. gpd plant which may be expanded later to 100,000 U.S. gpd. These will be the largest reverse osmosis plants to be constructed to date.

An important advantage in the reverse osmosis process is the fact that no phase change is involved and energy costs are low on that account. The advantage tends to be offset by the short life and high cost of membranes, which accounts for about two-thirds of the cost of the product water.

It is obvious that reverse osmosis, like electrodialysis, is likely to find most favor in regions where electric energy is cheaper than low-grade steam. It has been suggested that treatment of highly saline source waters in such areas might be effected most economically by combining the two processes. The feed water would be reduced to about 2000 ppm by reverse osmosis and then passed to an electrodialysis stack for further refinement to 500 ppm product water.

A modification of the reverse osmosis process, called Vapor Gap Reverse Osmosis, is now under study. It differs from normal reverse osmosis in that the membrane is permeable to water vapor but not to salt ions or to water in the liquid phase. Flow of vapor through the membrane is induced by raising the pressure on the saline solution so as to increase its vapor pressure above that of the fresh water. In a sea-water fresh-water system, vapor pressure equilibrium is established when the sea water is at 25 atmospheres pressure and the fresh water at about 1 atmosphere. An increase in pressure on the sea water induces flow of vapor to the other side of the membrane, where it condenses on the pure water and is drawn off as product water. Much of the current research is being directed at the development of satisfactory membranes.

Desalination by Freezing

Desalination by freezing has been under intensive study and has now reached the stage where units of up to 60,000 U.S. gpd capacity are being operated in pilot or field test plants. In principle, the saline feed water is cooled below freezing temperature to produce a slurry of brine and ice crystals which is pumped to a separator where the ice crystals are removed. Subsequently, the crystals are washed and melted to form the product water (Figure 10). The freezing may be accomplished by the direct evaporation of the water in the brine or, more commonly, by the evaporation of a refrigerant mixed with the feed water. The refrigerant vapor is drawn off from the freezer and condensed, giving up its heat of vaporization in melting the ice crystals, and it is then re-cycled.

The freezing techniques have an inherent advantage over distillation methods in that the phase changes (liquid to ice to liquid) involve energy transfers of only 144 BTU's per lb., the latent heat of fusion of water, as opposed to the 971 BTU's per lb. required for vaporization. Other advantages are the elimination of metal heat-transfer surfaces, except in feed water pre-cooling exchangers, and the elimination of serious corrosion and scaling tendencies because of the low operating temperatures. Heat losses from the system are minimized for the same reason.

The freezing process is applicable to all saline waters, including sea water, and it may be preferable to electrodialysis for water of high salt content. Like electrodialysis, it is likely to be considered where hydro-electric energy is cheaper than energy from fossil fuels. At its present stage of development, it is probably not competitive with modern distillation methods in plants ranging above 1 million U.S. gpd capacity.

A typical application of the freezing method is a 55,000 U.S. gpd test plant developed by the U.S. Office of Saline Water at St. Petersburg, Florida. Sea water is introduced to a freezer, where it is mixed with a butane refrigerant in the liquid phase. The butane evaporates, absorbing its heat of vaporization from the water and inducing formation of ice crystals. The resulting slurry of brine and ice is pumped from the freezer to a washer-melter employing the principle of column separation. Ice crystals float to the top of the column, where they are washed by a spray of fresh water and skimmed off into melting chambers surrounding the column. The butane vapor is drawn off the top of the freezer and compressed until its temperature exceeds the melting point of ice. It is then introduced to the melting chambers where it condenses in contact

with the ice and induces melting of the crystals. Brine is drained off midway up the separating column and residual butane decanted from it. Some of the brine is then re-cycled and the rest rejected through a feed water pre-cooler. A mixture of butane refrigerant and fresh water is pumped from the melting chambers to a decanter where the butane is drawn off the top, passed to a compressor and then re-cycled to the freezer. The product water is drained from the bottom of the decanter and passed through a feed water pre-cooler as it emerges from the system. A small quantity of the product water is diverted to the top of the separator to provide the spray for washing the ice crystals.

The freezing process is being used also in a 250,000 U.S. gpd plant at Wrightsville Beach, North Carolina. This is one of the five larger desalination demonstration plants being operated under the direction of the U.S. Office of Saline Water. Significant features of the plant are the use of a secondary refrigerant and centrifugal separation of ice crystals.

A 240,000 U.S. gpd plant made up of four 60,000 U.S. gpd units is now in operation in Eilat, Israel. Potable water of 200 to 300 ppm dissolved salts is being produced from 38,000 ppm water drawn from wells
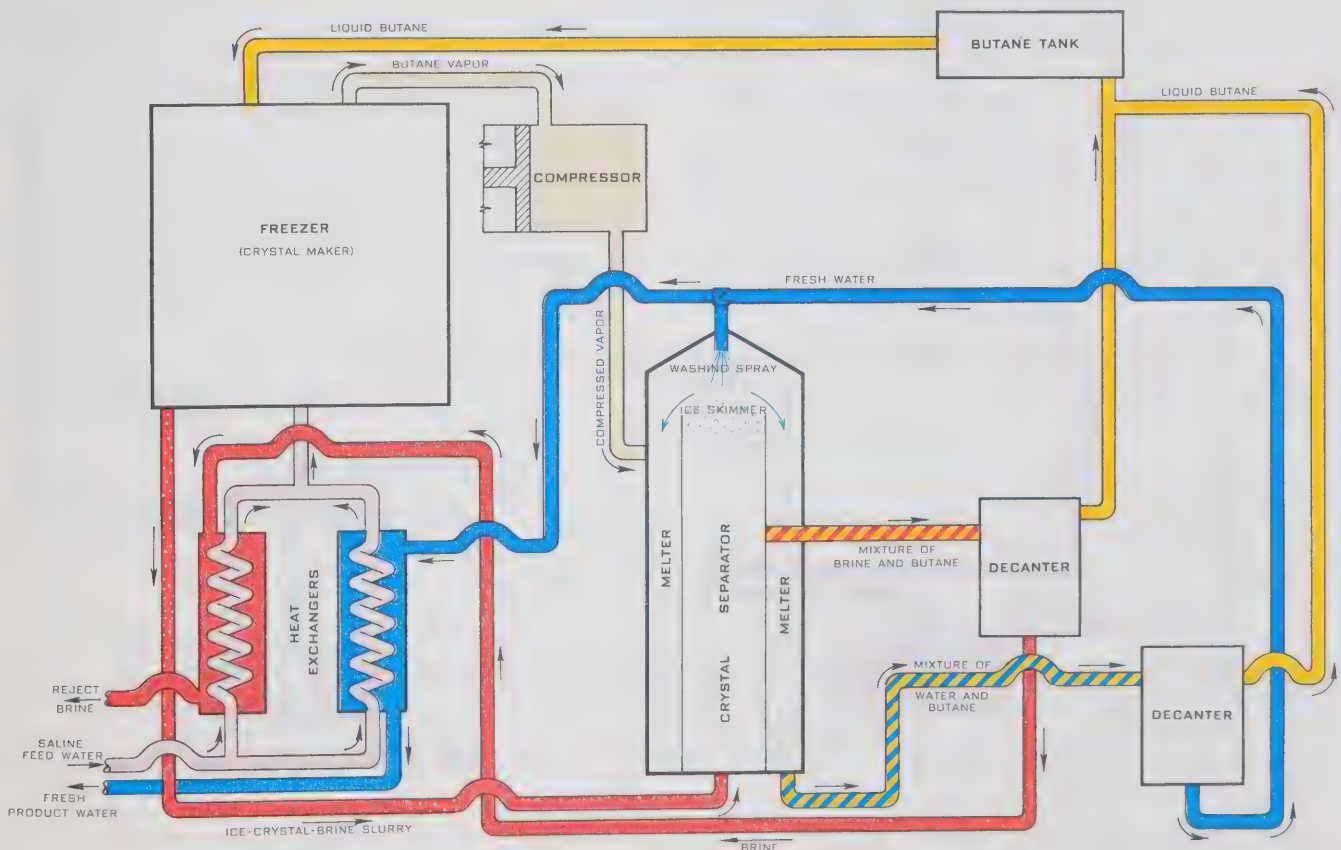
drilled in the beach on an arm of the Red Sea. The process employs direct freezing by evaporation in a vacuum. The vapor is drawn off, compressed and introduced to a "melter" where the condensation melts the ice crystals in the brine-crystal slurry from the freezer. The process is sometimes called the "vacuum freezing, vapor compression" process.

There are other test plants in which many modifications of the technique are being investigated. Refrigerants other than butane, the use of centrifuges rather than separation columns to extract the ice crystals from the brine, a two-stage freezing and separation process, and many other such variations are receiving attention. Development of efficient techniques for washing the ice crystals is one of the major problems.

Other Desalination Processes

ION EXCHANGE:

Desalting may be effected by passing saline water through beds composed of ion-exchanging resins. The feed water stream goes first to a bed of resin which has the property of absorbing the positive metal ions

FIGURE 10. FREEZING PROCESS USING BUTANE REFRIGERANT

of the dissolved salts and replacing them in the solution with positive hydrogen ions. Leaving this bed as an acidic solution containing positive hydrogen ions and the negative ions of the dissolved salts, the stream is passed to a second bed composed of resin with an affinity for negative salt ions. Here hydroxyl ions are given up to the stream and the negative ions of the salts absorbed. The stream emerges from the second bed stripped of its salt ions and containing only the hydrogen and hydroxyl ions in the form of chemically pure water.

The activity of the resins continues only until their respective hydrogen and hydroxyl ions have been completely replaced. Regeneration may be effected by passing a suitable acid through the first bed and a caustic solution through the second. Ion exchange occurs again, the hydrogen and hydroxyl ions of the washes being absorbed by the resins and the salt ions from the resins going into solution to form saline effluents which are run off to waste. For example, in desalting sea water the first bed absorbs the sodium ions and the second bed the chloride ions. Sulphuric acid and caustic soda are used as the regenerating agents. During regeneration, the resin beds absorb hydrogen and hydroxyl ions and the effluents are sodium sulphate from the first bed and sodium chloride from the second.

As the ion-exchange process consumes acids and bases in the regenerating sequence, each in the amount of the chemical equivalent of the salt in the feed water, it is quite impractical for large-scale, sea water applications. It may be feasible for desalting brackish waters with salt content below 3,000 ppm, perhaps as one stage in a combination process with electrodialysis or reverse osmosis. It is sometimes used in pre-treatment of feed water for distillation processes, as a means of controlling scale. In this application, it is not used to de-mineralize the water but rather to replace the salts which cause hardness.

HYDRATE PROCESS:

Some hydrocarbons combine with water to form solid "hydrate" crystals which are insoluble in water. This property leads to still another desalination process which has reached the pilot plant stage in current research (Figure 11). The hydrate crystals are formed

FIGURE 11. HYDRATE PROCESS
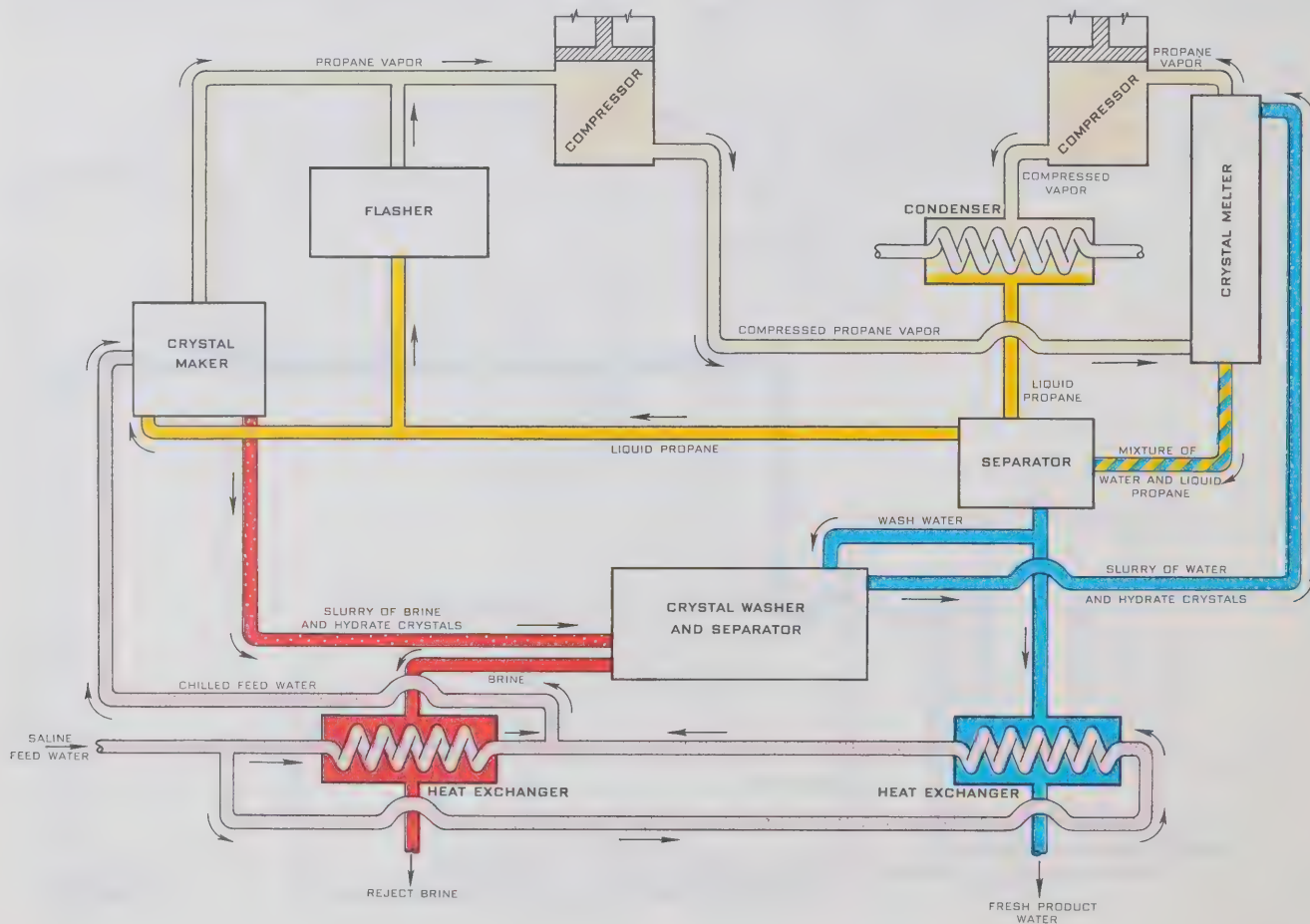
in a mixture of feed water and a hydrating agent such as propane. The resulting slurry is pumped to a separator where the crystals are extracted from the brine and washed. They are then passed to a melting unit where the hydrating agent is removed by vaporizing it at temperatures well below the boiling point of water, leaving the required product water. After leaving the melter, the vaporized hydrating agent is condensed and re-cycled to the "crystal-maker" unit. Heat for the crystal melter is obtained by compressing some of the hydrating agent in the vapor phase and then condensing it in the melter. The vapor is produced in a flash chamber, through which some of the cycling hydrating agent is by-passed, and is also picked up at various other points in the system where flash occurs spontaneously.

Pilot plants in the range of 10,000 to 20,000 U.S. gpd are now in operation. Several hydrating agents are being investigated, including propane, carbon dioxide and compounds of chlorine, fluorine and methane. Operating temperatures, separation methods, and other aspects are still under study and the process is still considered to be in the experimental stage.

SOLVENT EXTRACTION:

Fresh water may be extracted from a saline solution by the use of certain hydrocarbon solvents which have high selectivity in dissolving water and not salt, and whose ability to dissolve water is highly sensitive to temperature. The solvent and feed water are mixed at a temperature where water solubility is greatest and an extract of solvent and fresh water separated from the brine by decanting. The extract is heated to a temperature where water solubility is low, which results in precipitation of some fresh water from the extract. After the fresh water is decanted, the remaining extract is re-cycled to the feed water mixer, passing through a heat exchanger where its residual heat is recovered in raising the temperature of the extract leaving the mixer.

Although the process has no large-scale application as yet, a small 200 U.S. gpd pilot plant has produced potable water successfully from a 5000 ppm source water. The plant consumes relatively small quantities of input energy in raising the temperature of the extract by 30°F. This relatively low energy consumption has led to consideration of solar energy as a source of driving power and a solar heater has been developed which is capable of supplying the required heat for the pilot plant. Economic feasibility of the arrangement has not yet been demonstrated.

A more sophisticated system has been suggested in which a mixture of a solvent and a volatile liquid such as propane would meet a counter-flowing stream of saline feed water in a series of flash-evaporation chambers. The propane would flash into vapor, cooling the solvent and increasing its solubility to water. A solvent-water extract would emerge from the evaporating zone in the last chamber and be re-cycled through condensing zones in the chambers. Here the propane vapor would be condensed on the extract, thus raising the temperature of the extract and precipitating some fresh water. Product water would be decanted from the solvent-propane-water stream emerging from the condensing zones. The remaining solvent-propane mixture would then be pre-heated and returned to the evaporation zones in the chambers to mix with incoming feed water. Here the propane would again flash into vapor to initiate the new cycle. The process has not been demonstrated experimentally but preliminary studies have indicated its feasibility.

CURRENT DESALINATION RESEARCH

The basic principles now employed in the various water desalination processes have been known for some time. Spectacular breakthroughs into entirely new areas are not anticipated, despite the intensive research now under way. What may be expected is continuous improvement in techniques and equipment and a consequent progressive lowering of unit costs as more is learned about the interrelated behaviour of water, salts, solvents, metals and membranes.

The following brief review of a few of the activities connected with the current research program sponsored by the U.S. Office of Saline Water is indicative of the kind of studies that are now being presented in many countries:

- Studies of the properties of water and its behaviour as a solvent and a solute.
- Studies of the properties and behaviour of salt solutions.
- Research into the principles involved in transfer of energy or mass, adjacent to or across phase boundaries.
- Study of the passage of water and salt ions through membranes and of the deposition of salt, scale and slime on membranes.
- Development of improved ion-selective and reverse osmosis membranes and research into factors affecting membrane durability.
- Research into improved methods of crystal formation and separation in freezing and hydrate processes.
- Research related to reduction of corrosion. Examples are studies of corrosion-resistant metal alloys, coatings for metal heat-transfer surfaces, electrolytic and acid controls, use of concrete for evaporation chambers.

- Research relative to reduction of scale deposition on heat transfer surfaces.
- Investigation of economic means for recovery of minerals from brine as a by-product of desalination.

Meanwhile, applied research in pilot and demonstration plants and in a few commercial installations is being directed toward economic solutions to engineering problems of design, operation and maintenance of desalting systems. Experience at these plants is providing a realistic basis for economic and technical feasibility studies of much larger plants with capacities in the range of 10 million to 200 million U.S. gallons per day.

DUAL-PURPOSE PLANTS

The combination of power generation and desalting functions in a dual-purpose plant is a recent development which will play an important role in future desalination technology. In a dual-purpose installation, low-grade exhaust steam from a steam turbine (or hot gas from diesel engines or gas turbines) is used as the input energy to a desalter employing a distillation process. In a sense, the distiller becomes the condenser for the steam turbine.

The principal advantage in the dual arrangement is that it uses in the desalter the low-temperature heat of the exhaust steam which would otherwise be wasted to the ambient. As the exhaust steam must be delivered to the desalter at a pressure higher than the usual condensing pressure, the dual-purpose plant produces less power per pound of steam than the single-purpose turbogenerator. It may, however, secure an overall advantage from its fuller use of the total heat of the steam.

The dual system has another potential advantage in the opportunity afforded to improve the load factor of the steam source, providing the plant design can effectively balance the power and water requirements. As the load curves for these demands do not conform exactly, the system must have sufficient flexibility in the "water to power" ratio to overcome the problem. One of two basic situations will be confronted to some degree in any specific application. In one case, the generator may be a unit in a large grid which can absorb the entire power output at all times. The desalter can then be operated at a high load factor and power output geared to water demand. In the other case, the requirement for exhaust steam to satisfy water demand may call for a turbo-generator whose capacity represents a substantial fraction of the installed capacity of the region. The power output must then conform to the regional load curve and water output will fluctuate with power demand. In either case, facilities for water storage will be required and also some provision for operation of the turbo-generator during periods of shutdown of the desalter.

Either of two basic design principles, called the "back pressure cycle" and "pass-out condensing cycle" respectively, may be employed in a dual-purpose installation. In the back-pressure cycle, all the exhaust steam is fed to the desalter through a reducing valve, which controls the temperature and pressures in both the turbine and the desalter. The output of power and water are intimately associated. Where necessary, flexibility may be obtained by introducing a line to by-pass the turbine and feed steam directly from the boiler to the desalter during periods of low power demand. The sacrifice of energy production, however, makes the fresh water produced by the by-pass steam more expensive. In the pass-out condensing cycle, a high-pressure and a low-pressure turbine are provided and the steam for the desalter is bled off between the two sections. In this case, the high-pressure unit performs in the manner of the usual single-purpose turbine, accepting steam from the boiler on command from a governor. A valve in front of the low-pressure stage controls the pressure of the steam entering the desalter and accepts the excess for power production in the low-pressure turbine.

Another system has been suggested in which high thermal efficiency and flexibility would be achieved by the use of two different desalting processes. A loop consisting of a boiler, steam turbine and distiller would be operated continuously at full load. Variations in the power demand would be smoothed out by directing excess electric power to an electrodialysis unit.

It must be emphasized that the feasibility of a dual-purpose system cannot be assessed nor the most economic design selected without a detailed analysis of the many technical, economic and social factors that come into play. It is possible that, in some circumstances, a separate desalting unit supplied by energy purchased on the open market would prove to be a more economic solution.

Although the term dual-purpose desalination is usually employed in reference to plants producing power and fresh water, it might be applied to an installation where by-products other than power are marketed. An example is found in Kuwait where salt, chlorine, hydrochloric acid and caustic soda are being processed as by-products from a desalter. It is possible that a desalter using the freezing technique might also market ice for refrigeration, cold storage capacity or cool water for air conditioning.

DESALINATION USING NUCLEAR ENERGY

The possibility of using a nuclear reactor as the heat source for a desalination process has been under study for the past few years. Some general principles concerning the feasibility of such a plant are now generally accepted. Nuclear fuel is not likely to be able to compete with fossil fuels in single-purpose installations, chiefly because the nuclear reactor provides high-grade steam, which is unsuitable for desalters on account of the scale and corrosion problems associated with the high temperatures. In dual-purpose plants for production of both water and power, where high-grade steam is expanded in a turbine before entry to the desalter, nuclear fuel is feasible but not economically competitive in small installations because of the large investment in the reactor. Nuclear fuel becomes more competitive as the capacity of the desalter increases and is definitely more economic in very large dual-purpose plants because of the lower unit cost of fuel. This presupposes, of course, that there is a market for large quantities of power.

There are no nuclear-fueled desalters in operation yet but plans are well advanced for a plant of 150 million U.S. gpd capacity to be constructed as a joint undertaking by the U.S. Office of Saline Water, the Metropolitan Water District of Southern California and electrical utilities in the California area. The desalter will be operated in conjunction with a 1500 MW nuclear power plant. Present plans call for product water to be available by 1971 at an estimated cost of 21.9 cents per thousand gallons. Several other countries, including several of the Mediterranean countries, are considering the installation of nuclear, dual-purpose plants.

Most of the research into nuclear-fueled plants has been directed toward light-water reactors but it has been contended that a heavy-water reactor, such as the CANDU reactor developed in Canada, may have economic advantages. Canada may have an opportunity to capitalize on its know-how in this field in the design and export of desalting units powered by CANDU-type reactors.

SOME ASPECTS OF THE ECONOMICS OF WATER DESALINATION

The cost of producing fresh water from a highly saline source has been sharply reduced since intensive research and experimentation in large-scale desalination got under way in 1952. One thousand U.S. gallons of desalted water, which would have cost over $4.00 in 1952, can now be produced for about $1.00 to $1.50, or even less if the source water is a brackish type

with relatively low salt content. Meanwhile, several alternative desalting processes have been brought through the laboratory and pilot plant stages to demonstration or commercial installations with capacities of individual units ranging up to 1 million U.S. gallons per day (Table 2). Plants of ten to one hundred times this capacity, now in the design or construction phases, will be producing a considerably cheaper product water within the next five years, the estimates in some cases running as low as 22 cents per 1000 U.S. gallons for nuclear-fueled, dual-purpose plants (see Figure 12) *.

Clearly, desalinated water can no longer be lightly dismissed as an expensive luxury in planning for future use of water resources. It is already competitive in regions where fresh water from natural sources is in extremely short supply. In other areas, as expanding populations and rising living standards impose increasing demands on existing natural supplies while the unit cost of fresh water from large desalters declines, the desalted water will become progressively more competitive. It may be expected that desalination will provide an economic solution to a growing number and variety of water supply problems.

The total costs of desalted water at the plant may be conveniently separated into two components: fixed cost and variable costs. The fixed costs are those which do not change with the daily production of fresh water and are principally the interest on the original investment and the depreciation charges on the equipment. Maintenance, land rental (if applicable) and insurance are also included. The fixed costs may approximate half the total cost of the operation and thus interest and depreciation rates are strongly reflected in the reported costs of product water from existing installations.

The variable costs are those which increase with the quantity of the product, the most important being the cost of input energy. Generally speaking, the total energy consumption in a specific operation varies directly with the quantity of fresh water produced. It is also governed by the temperature of the source water and, in some processes, by the salt content of both the source and product waters.

Escalation in the size of the plant does not have a significant effect on the energy required for a gallon of fresh water but it may reduce unit capital cost and thus effect an over-all reduction in the cost of the product water.

*In the interval since Figure 12 was prepared, an estimate of 36 cents per 1000 U.S. gallons for product water from a single-purpose, MEMS flash evaporation plant of 50 million gpd capacity has been submitted to OSW by the Worthington Corporation.

TABLE 2

Examples of desalination plants now in operation

Process	Location	Rated capacity of plant US gpd	No. of units	Rated capacity of units US gpd	Salinity of source water ppm	Salinity of product water ppm	Remarks
Long Tube Vertical Distillation (LTV)	Freeport, Texas	1,000,000	1	1,000,000	Sea water 35,000	50 ppm (max.)	Demonstration Plant of OSW*
Submerged Tube Distillation	Aruba, Netherlands Antilles	2,600,000	5	530,000		5	Dual purpose installation
	Curaçao, Netherlands Antilles	1,000,000	2	500,000	41,200	120 (max.)	Dual purpose installation
	Kuwait	1,200,000	10	120,000	45,000	50	Dual purpose installation
Flash Evaporation	Guantanamo Bay Cuba	2,250,000	3	750,000	Sea water		Dual purpose installation
	Curaçao, Netherlands Antilles	3,000,000	2	1,500,000	Sea water		Dual purpose installation
	Kuwait	2,000,000	2	1,000,000	45,000	50	Dual purpose installation
Vapor Compression	Roswell, New Mexico	1,000,000	1	1,000,000	24,470	50 (max.)	Demonstration Plant of OSW
Electrodialysis	Webster, South Dakota	250,000	1	250,000	1,800	275 (max.)	Demonstration Plant of OSW
	Buckeye, Arizona	650,000	1	650,000	2,100	500	Municipally owned plant
	Welkom, South Africa	2,880,000	1	2,880,000	3,000	500	Plant now only on "stand by" Status
Freezing	Wrightsville Beach, North Carolina	250,000	1	250,000			Demonstration Plant of OSW
	Eilat, Israel	240,000	4	60,000	38,000	500 (max.)	

*OSW — Office of Saline Water of the United States Department of the Interior.

In regions which have a plentiful supply of cheap fossil fuel, thermal distillation is likely to be preferred because of the relatively low unit capital cost. Where low-grade steam is expensive, processes such as freezing, vapor compression or electrodialysis may be more attractive despite higher capital costs, especially if hydro-electric plants are the principal source of energy. On the other hand, factors other than energy cost may restrict the choice of processes and hence influence the cost of the product water. Electrodialysis, for example, may be the most economic solution to treatment of brackish water but it would be quite uneconomic when dealing with sea water, and it may not be feasible for brackish waters with high concentrations of scale-forming salts.

The economics of desalination becomes much more complex in the case of dual-purpose plants, the cost of the product water being intimately associated with the production and sale of power. Separation of the costs attributable to the power and water outputs in any individual case is difficult and may involve somewhat arbitrary assumptions. Moreover, the design of the plant will be strongly influenced by the relative size of the power and water demands and the shape of their respective load curves. These and other factors have a combined effect on the cost of the product water which will not be duplicated in any other place and time. The data from feasibility studies for a specific application, although they may be indicative of the order of magnitude of costs of fresh water from dual-purpose plants, are not readily transferable to another site on this account.

With these reservations in mind, it is of interest to note that feasibility studies on hypothetical, single and dual-purpose plants, both nuclear and fossil-fuelled, with capacities ranging between 15 million and

260 million U.S. gpd, indicate that sea water could now be desalinated for "at plant" costs in the order of 21 cents to 70 cents per 1000 U.S. gallons, depending upon many relevant design and operating criteria. Transportation and distribution costs must be added to obtain a realistic estimate of cost of product water delivered to the consumer. Although these added costs would vary with such factors as pumping lifts and distance from plant to consumer, estimates of 15 cents/1000 gallons/100 miles for transportation and 35 cents/1000 gallons for distribution would be reasonable in making rough comparisons with present costs of fresh water from natural sources.

The technical feasibility of large-scale desalination by one of several different processes has now been demonstrated. Its application to any specific water supply problem hinges on the availability and cost of alternative supplies of fresh water. In many communities in the United States and Canada, domestic and industrial water is still supplied on a flat rate and realistic cost data are not readily available. In those communities where water is sold on the basis of consumption, prices to the consumer are in the order of 10 cents to 70 cents per 1000 U.S. gallons. Some examples are cited in Table 3.

It is obvious from the figures presented that the existing desalination plants with capacities ranging up to 1 million U.S. gpd would be unable, at the present stage of development, to supply consumers in most communities in Canada or the United States with potable water from a highly saline source at prices they have been accustomed to pay. This is even more evident when it is considered that the given costs of desalinated water do not include transportation and dis-
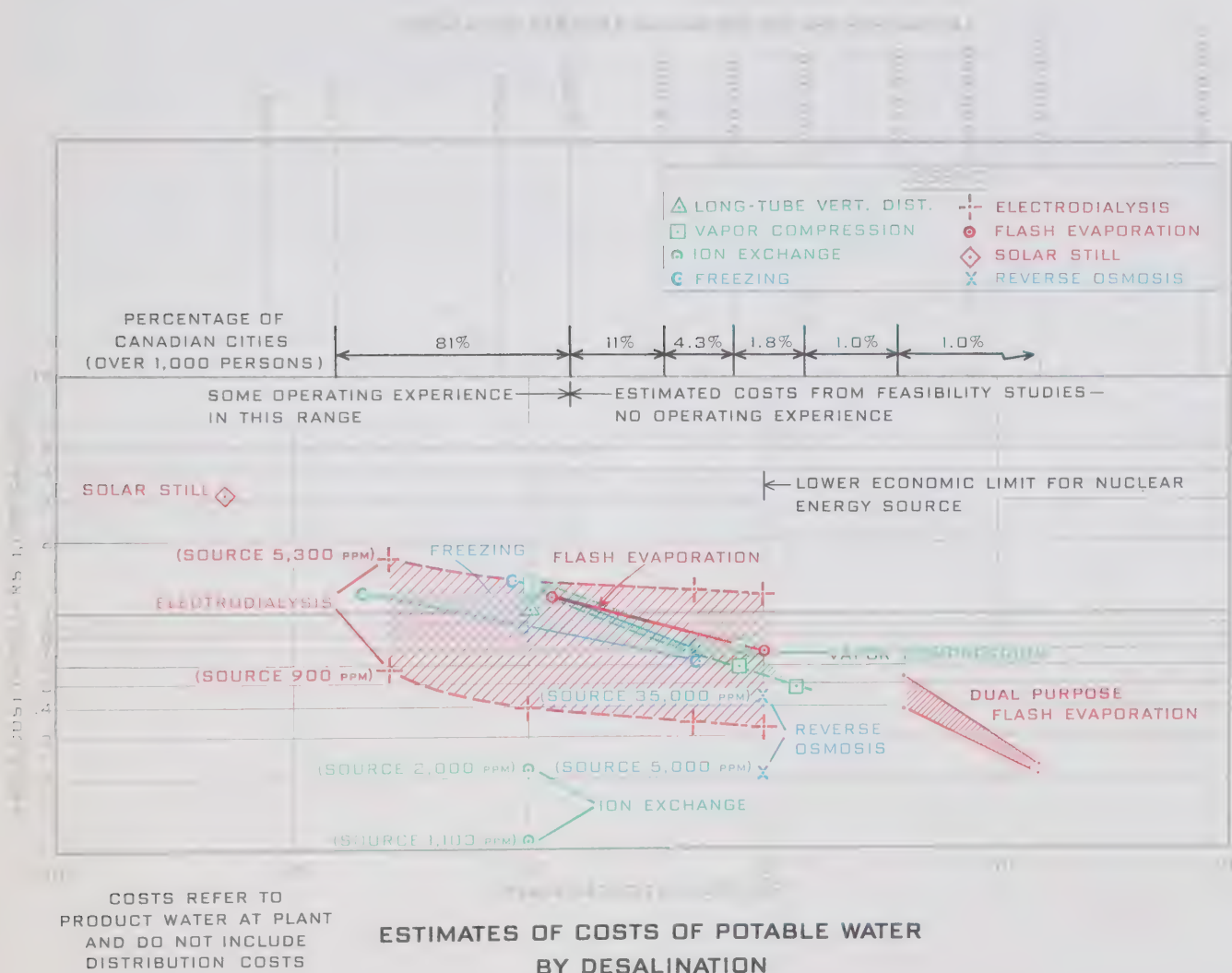

FIGURE 12. COST OF WATER DESALINATION

TABLE 3

Examples of Present Municipal Water Prices

City	Purpose	Price per 1000 Imp. gallons in Canada	Equivalent price per 1000 U.S. gallons in U.S. funds
Ottawa	Domestic	$0.34	$0.26
Regina	Domestic · Commercial	0.61	0.47
	(Treatment)	(0.165)	(0.125)
	(Distribution)	(0.28)	(0.215)
	Total	0.45	0.34
Calgary	Domestic		
	1st 1000	0.62	0.48
	Next 1000	0.56	0.43
	Next 20,000	0.50	0.38
Washington, D.C.			0.27

tribution costs. On the other hand, water supply systems are usually subsidized to some degree from other sources of revenue and the present gap between the real costs of natural and desalinated water may not be as high as appears at first glance.

Indeed, where the only readily available source of supply is a brackish water with less than 3000 ppm dissolved salts, desalination may already be economic. This has been demonstrated in Buckeye, Arizona, where well water with about 2100 ppm salts is being refined to 500 ppm in an electrodialysis plant. A net saving to the community has been effected by a significant reduction in expenses for water softeners, soap, bottled water and replacement of corroded equipment.

Reclamation of liquid wastes in water-short areas of dense population is now being considered. Water becomes more saline with use and reclamation will involve removal of salts. Desalination of waste water, probably by an electrodialysis or ion-exchange process, may already be the most economic means of adding to the fresh water supply in some localities, its use in this way awaiting only public acceptance.

The feasibility studies mentioned previously have also shown that desalination of highly saline water is now competitive at some points on the North American continent, where fresh water is in short supply, where there is a saline source water close to the point of consumption and where demands are high enough to support plants of capacity in the tens or hundreds of millions of gallons per day. Plants of this type are now being considered for southern California, Florida and the New York City area. These are examples of situations in which large-scale desalination is likely to be most appealing at this time. They involve densely populated coastal areas capable of supporting such an operation, where increasing demands for natural fresh water would otherwise have to be satisfied by importing large quantities from some distance inland. Desalination will not only save the transportation expense but will also relieve the inland water for use nearer its point of origin.

With the exception of small truck-gardening uses in the humid areas, experience throughout Canada and the United States has shown that irrigation water must be supplied to the farm at prices in the order of 1.5 to 16 cents Canadian per 1000 Imp. gallons (U.S. equivalent 1 to 12 cents per 1000 U.S. gallons) depending upon the scale of the operation and the type of crops. This price in no way covers the full capital and operating costs and irrigators are heavily subsidized by public funds, the case for subsidies being based on the indirect benefits accruing to the community and the country at large. There is no indication that desalting costs will meet the economic figure for large-scale irrigation within the predictable future.

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